DME(10 TPD) Process Simulation for CO_2 Absorber & CO_2 Stripper Using PRO/II with PROVISION

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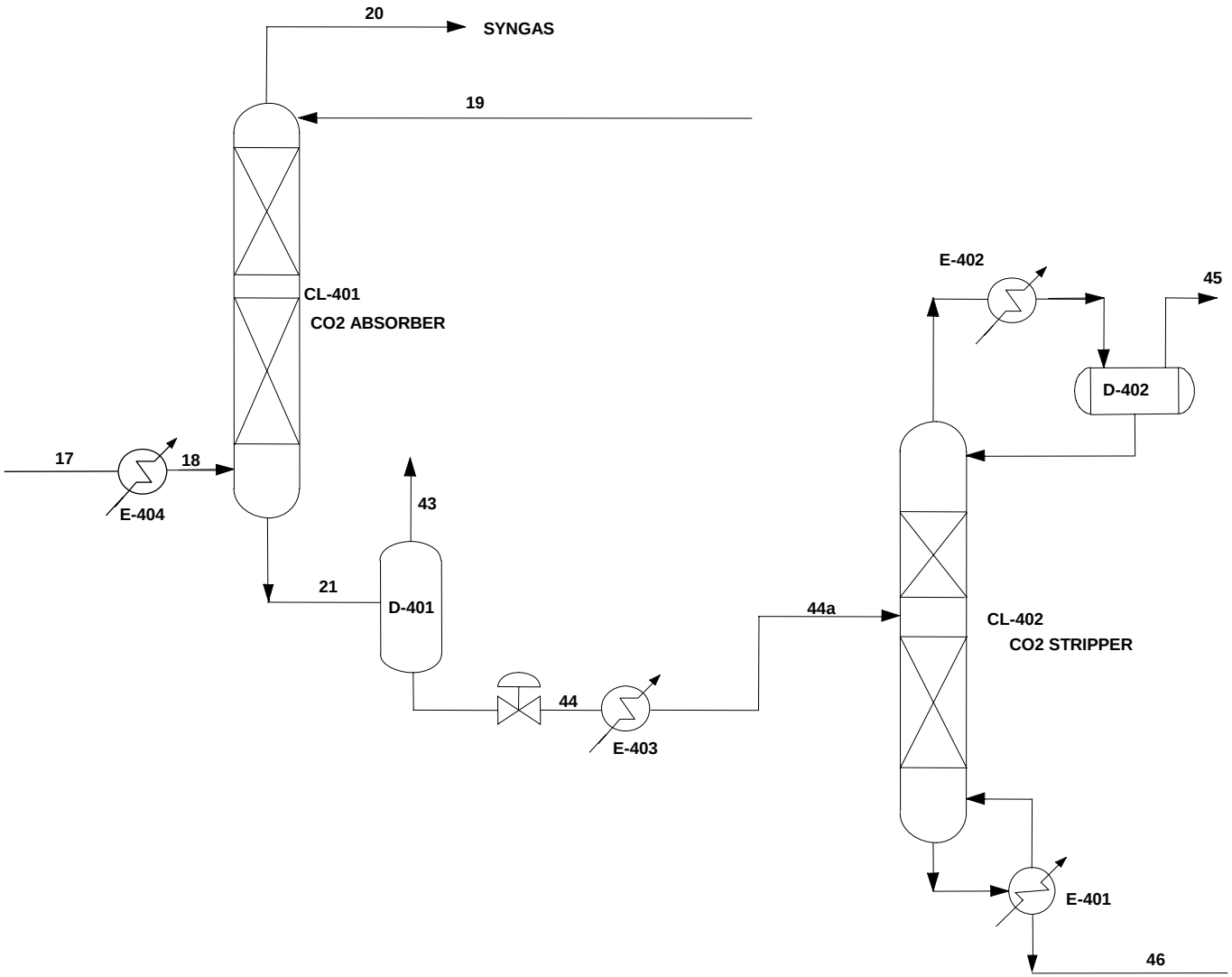
Overall Flow Sheet for DME Production Unit



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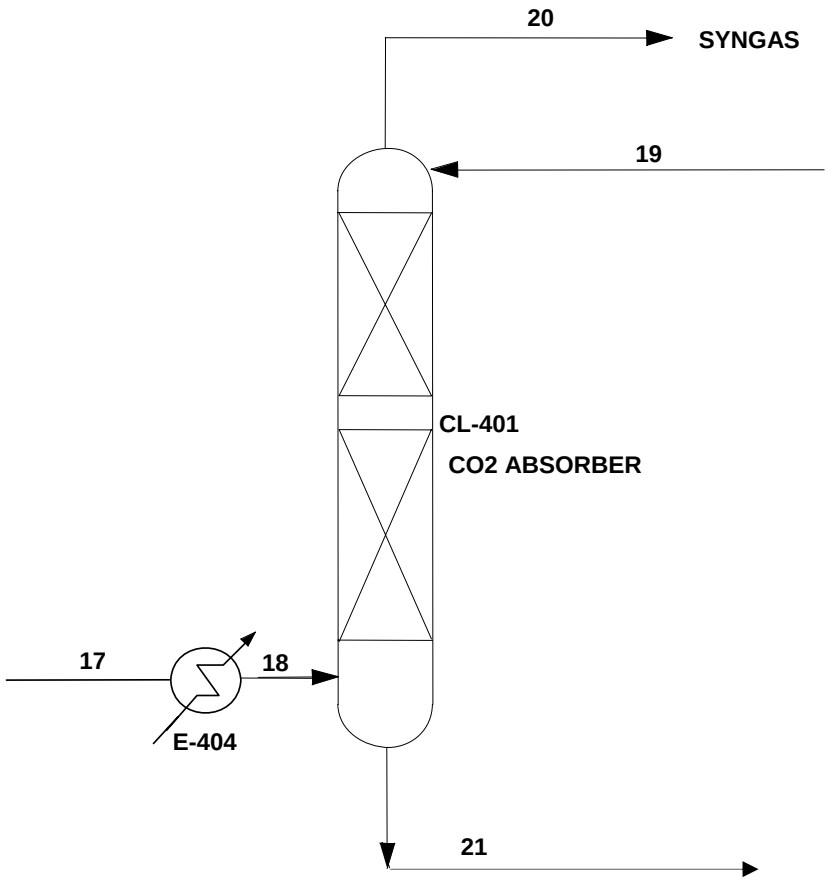
Flowsheet for Toluene Recovery Process

Unit	Description
CL-401	CO2 Absorber
CL-402	CO2 Stripper
D-401	CO2 Flash Vessel
D-402	CO2 Stripper OVHD Drum
E-401	CO2 Stripper Reboiler
E-402	CO2 Stripper OVHD Condenser
E-403	CO2 Stripper Preheater
Stream	**Description**
17	Reactor Outlet Stream
18	CO2 Absorber Feed Stream
19	Methanol Solvent Feed Stream
20	Synthetic Gas
21	Rich Solvent Stream
43	Flash Gas Stream
44	CO2 Stripper Feed Stream
45	CO2 Gas Stream
46	Lean Solvent

Overall Material Balance: Design Case

	17	18	19	20	21	44	45	46
Temperature (°C)	110.8	30.0	30.0	29.54	33.1	33.82	31.29	185.7
Pressure (kPa, abs)	6,000	6,000	6,000	6,000	6,000	3,000	3,000	3,000
Flow (k-mole/hr)	**96.9700**	**96.9700**	**329.8000**	**67.8100**	**358.9000**	**356.5**	**27.1**	**329.5**
Total Kg/hr	**2,194.0**	**2,194.0**	**10,570.0**	**984.0**	**11,780.0**	**11,710.0**	**1,159.0**	**10,550.0**
Component Molar Flow								
1. CO	33.6200	33.6200	0.0000	32.3000	1.5800	0.7490	0.7520	0.0000
2. H2O	0.1842	0.1842	0.0000	0.0000	0.1795	0.1783	0.0000	0.1977
3. CO2	26.5700	26.5700	0.0000	0.0610	26.5227	26.5700	25.6600	0.0000
4. H2	36.300	36.300	0.0000	35.2300	1.0800	0.3922	0.3760	0.0000
5. Methanol	0.0000	0.0000	329.8000	0.2577	329.4702	329.5000	0.2000	329.3000
6. DME	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
7. CH4	0.1842	0.1842	0.0000	0.1356	0.0359	0.0357	0.0433	0.0000
8. N2	0.1067	0.1067	0.0000	0.0949	0.0000	0.0000	0.008115	0.0000
9. O2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000

CL-401 (CO2 Absorber)



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CO2 Absorber Simulation (CL-401)

- Primary objective of the CO2 absorber is to absorb CO2 contained in the feed stream by contacting counter-currently with methanol solvent in an absorber.

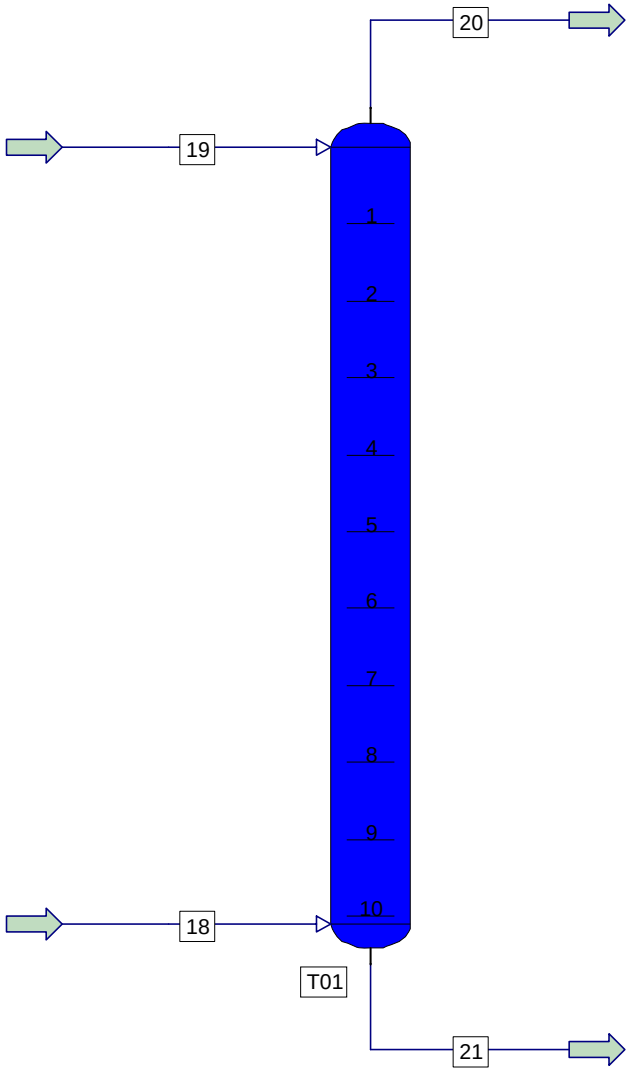
CO2 Absorber Simulation (CL-401): *Continued*

- Consider the following absorber column to absorb CO_2 contained in the feed stream using methanol as a solvent.
  - Feed1: Crude Feed (Refer to feedstock characterization)
  - Feed2: Methanol Solvent
    1) Solvent Feed Temperature: 30°C
    2) Methanol Purity: 100%
    2) Flowrate: 329.80 K-mole/hr
  - CO2 Absorber Column
    1) Number of Theoretical Stages: 10
    2) Column Type:
    3) Overall Tray Efficiencies: ?
    4) Feed Tray Location: 10
    5) Solvent Feed Tray Location: 1

DME Absorber Simulation *Continued*

- Selection of appropriate thermodynamic model for the simulation of DME absorber using methanol as a solvent is very important.
  - NRTL (Non Random Two Liquid) activity coefficient model was chosen to explain non-ideal phase behavior of liquid mixture between H_2O, DME, methanol and methanol.
  - Peng-Robinson equation of state method was used for the prediction of the vapor phase non-idealities since the system pressure is high.
  - Henry's law option was also selected for the calculation of non-condensible supercritical gases like H_2, CO, CO_2, CH_4 and N_2 in a liquid mixture, especially methanol solvent.

Flow Sheet Drawing Using PRO/II



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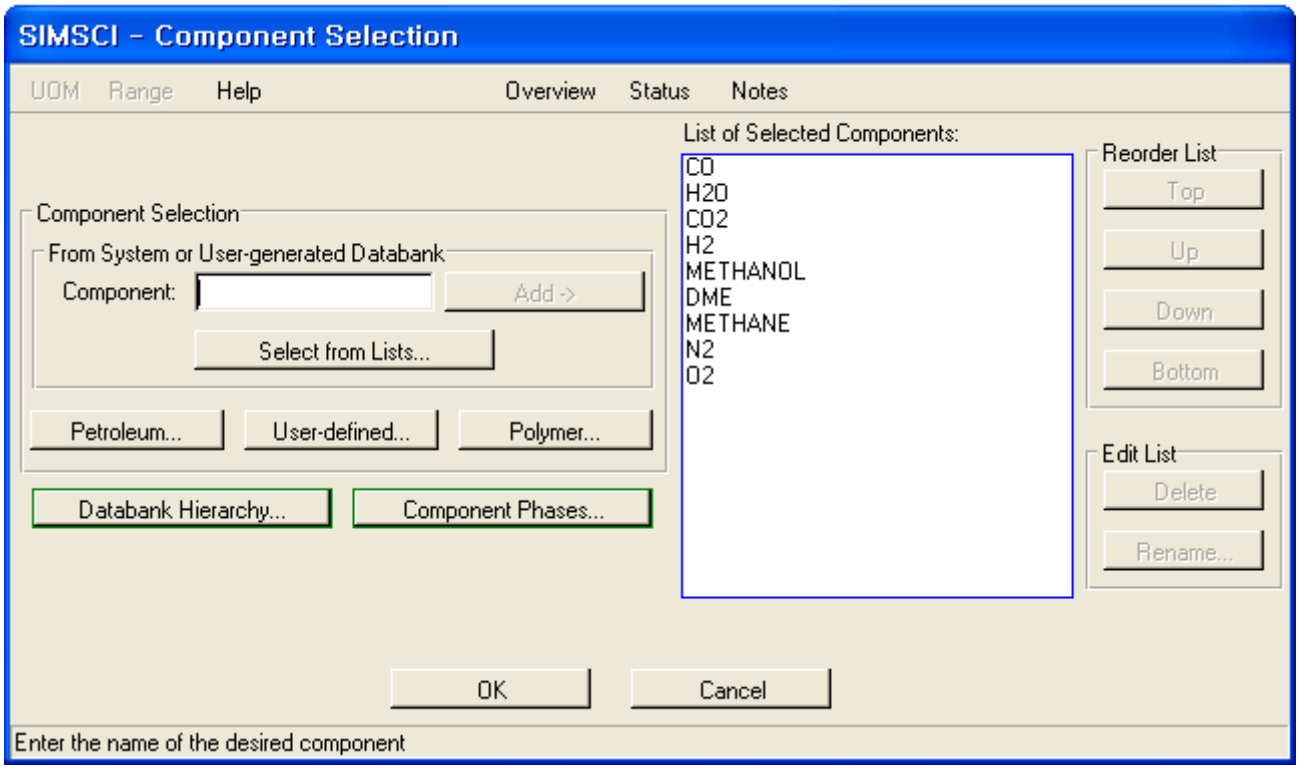
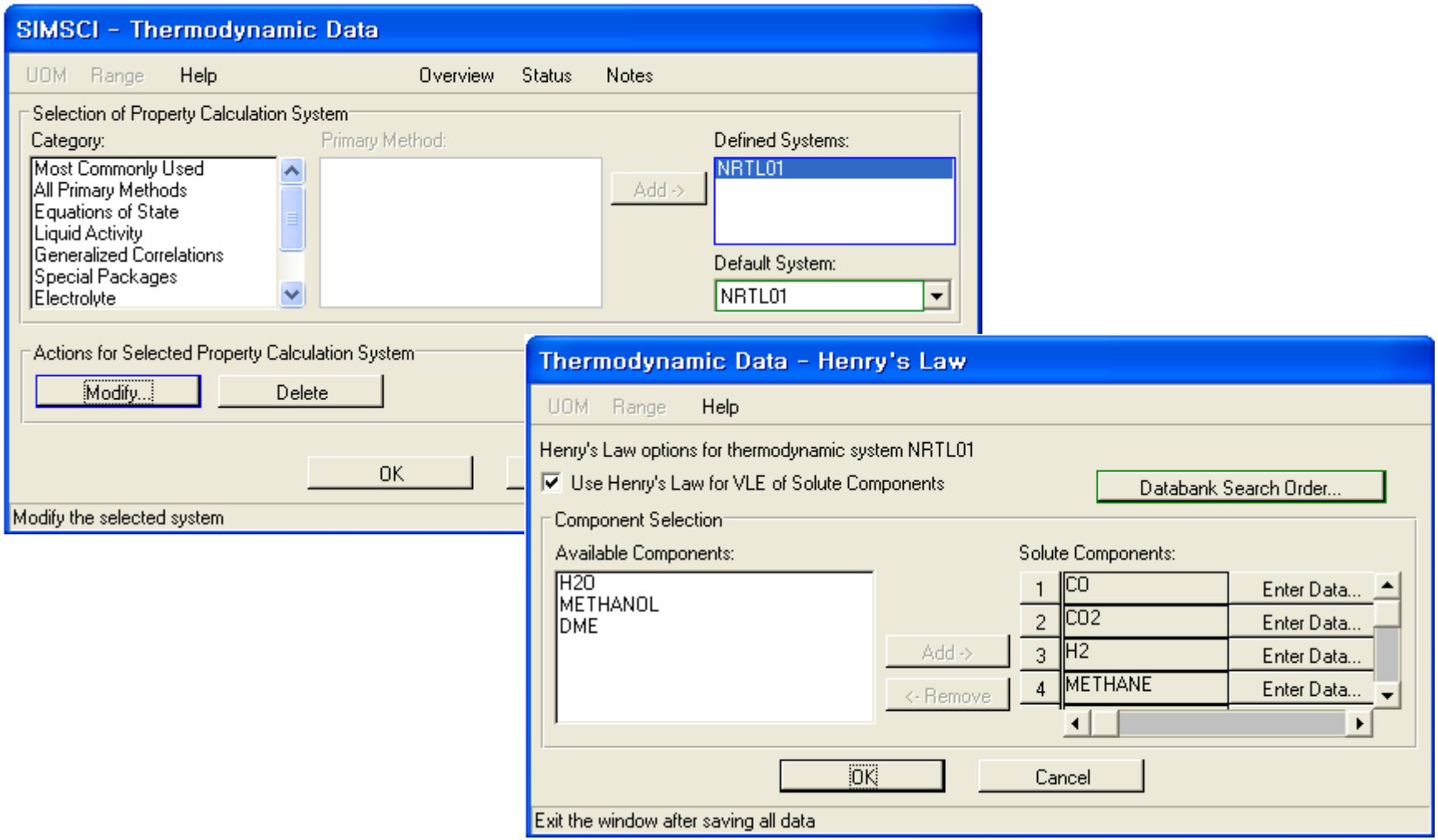
Component Data: Pure Component



Thermodynamic Data: NRTL with Henry + PR EOS



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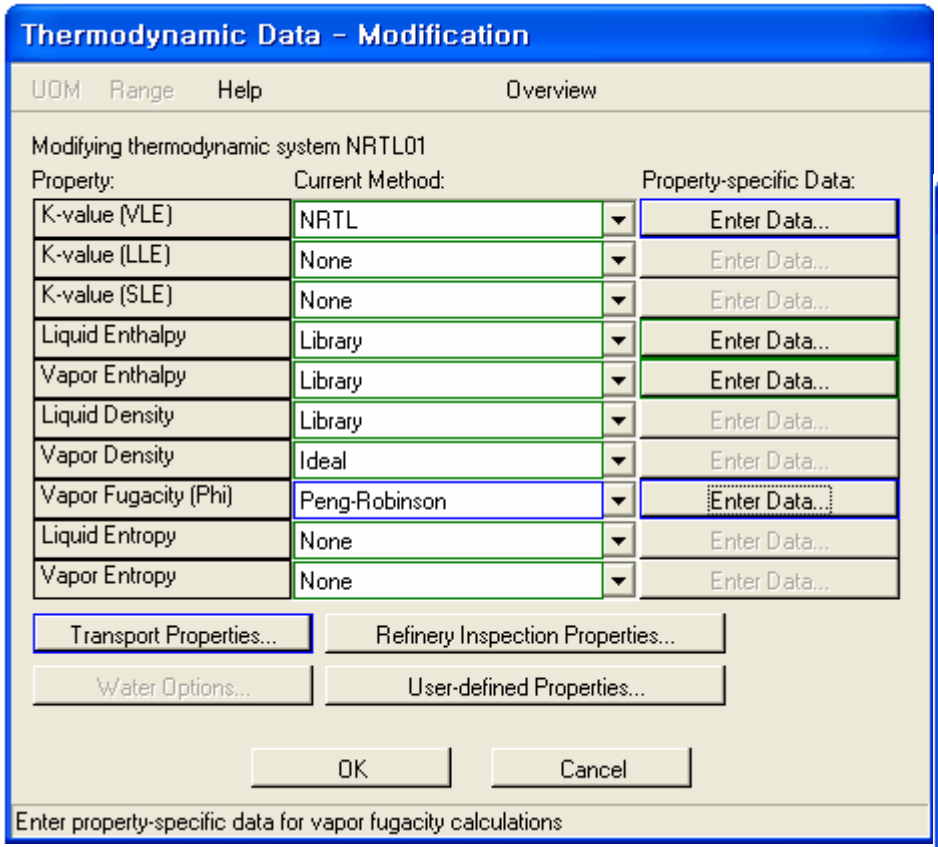

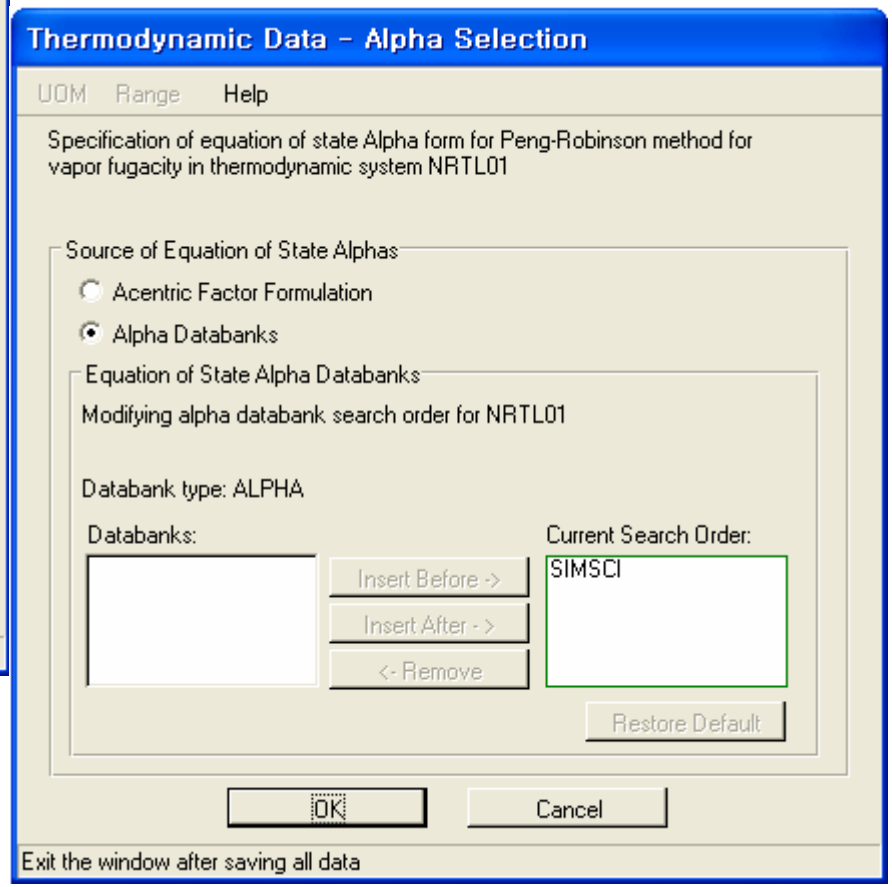

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NRTL(Non Random Two Liquid) Model

- NRTL. This model has up to 8 adjustable binary parameters that can be fitted to data.

$$\ln \gamma_i = \frac{\sum_j \tau_{ji} G_{ji} x_j}{\sum_k G_{ki} x_k} + \sum_j \frac{x_j G_{ij}}{\sum_k G_{kj} x_k} \left[\tau_{ij} - \frac{\sum_l x_l \tau_{lj} G_{lj}}{\sum_k G_{kj} x_k} \right]$$

$$\tau_{ij} = a_{ij} + \frac{b_{ij}}{T} + \frac{c_{ij}}{T^2}$$

$$G_{ij} = \exp\left[-\left(\alpha_{ij} + \beta_{ij} T \right) \tau_{ij} \right]$$

Henry's Law Option

- When using activity coefficient property methods for supercritical components, use Henry's law to better predict gas solubilities in the liquid phase.

$$K_i = \frac{y_i}{x_i} = \frac{H_{i,mix}}{P}$$

$H_{i,mix}$ is Henry's constant of component i in the mixture.

- H_i is calculated from temperature-dependent (and also pressure-dependent) Henry's constants for each solute-solvent pair.

Peng-Robinson Equation (1976)

- Peng-Robinson equation of state modified a functional form to fit better the liquid density so it is known to estimate well for midrange hydrocarbon (C_6 to C_{10}) systems.

$$P = \frac{RT}{V - b} - \frac{a_c(T_c) \cdot \alpha(T_r, \omega)}{V(V + b) + b(V - b)}$$

Peng-Robinson Equation (1976)

- Changed the cubic from Redlich-Kwong slightly.
- Changed the constants in Soave's alpha slightly.

$$\alpha_i(T) = \left[1 + m_i\left(1 - T_{ri}^{1/2}\right)\right]^2$$

$$m_i = 0.37464 + 1.54336\omega_i - 0.26992\omega_i^2$$

Peng-Robinson Equation (1976)

- The pure component a_c & b_c are found from:

$$a = 0.45724 \frac{R^2 T_c^2}{P_c} \qquad b = 0.07780 \frac{RT_c}{P_c}$$

- An alternative dimensionless form is given by:

$$P_r = \frac{3.2573\ T_r}{V_r - 0.2534} - \frac{4.8514\ \alpha}{V_r^2 + 0.5068\ V_r - 0.0642}$$

Critical Compressibility Factors

- Experimental values for critical compressibility:
  - 0.2880 for CH_4
  - 0.2840 for C_2H_6 &
  - 0.2800 for C_3H_8

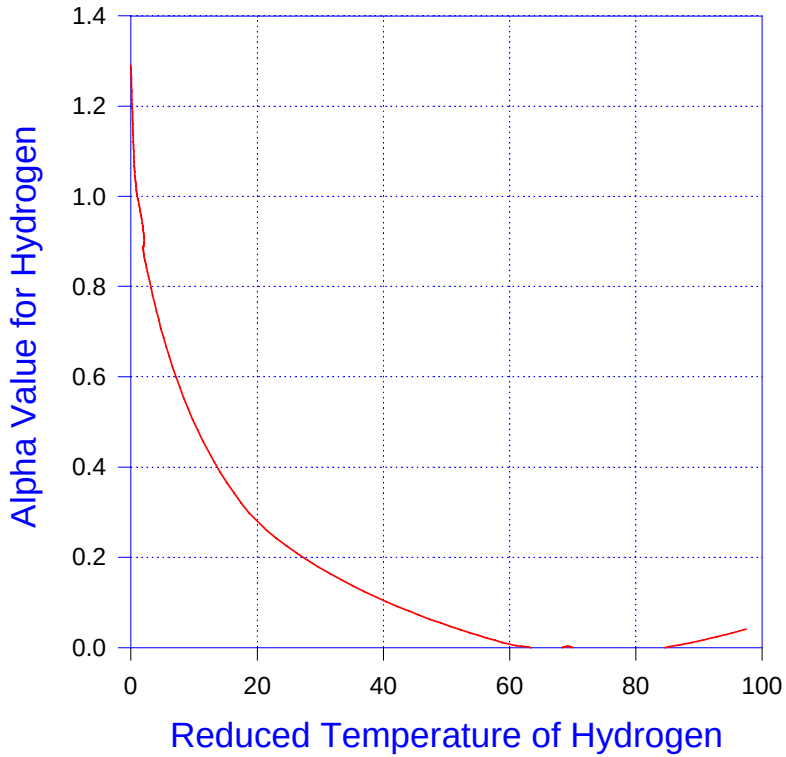
- PR and SRK equation of state are still used to estimate phase equilibria calculation for non-ideal systems since what is the most important thing in the design of chemical process is K-values, *not the liquid densities.*

Soave's Original Alpha form

- Good representation of liquid vapor pressure:

  "Proper alpha form"

- Soave's original alpha form is wrong since it increases again as reduced temperature of hydrogen, T_r approaches to infinity.



$$\alpha(T) = \left[1 + \left(0.37464 + 1.54226\,\omega_i - 0.26992\,\omega_i^2\right)\left(1 - \sqrt{T_r}\right)\right]^2 \quad (11)$$

Requirements for Alpha form

- Requirements for alpha form:
  - The α function must be finite and positive for all temperature.
  - The α function must equal unity at the critical point.
  - The α function must approach a zero value as the temperature approaches infinity.

- The trend from now is to set the coefficients of alpha function component dependently by regressing the experimental vapor pressure data vs. temperatures.

Several Alpha functions

$\alpha = \left[1 + C_1\left(1 - T_t^{0.5}\right)\right]^2$	Soave (1972)
$\alpha = \left[C_1 + C_2\left(1 - T_t^{C_3}\right)\right]^2$	Peng-Robinson (1980)
$\alpha = 1 + \left(1 - T_r\right)\left(C_1 + \frac{C_2}{T_r}\right)$	Soave (1979)
$\alpha = \exp\left[C_1\left(1 - T_r^{C_2}\right)\right]$	Boston-Mathias (1980)
$\alpha = T_r^{2(C_2-1)} \exp\left[C_1\left(1 - T_r^{2C_2}\right)\right]$	Twu (1988)
$\alpha = T_r^{C_3(C_2-1)} \exp\left[C_1\left(1 - T_r^{C_2C_3}\right)\right]$	Twu-Bluck-Cunningham (1990) (Recommended by SimSci)

New Alpha Form

- Since 1972, many alpha forms have been proposed, some better than others.
- PRO/II allows input of parameters for 11 different forms, including the SIMSCI (TBCC) alpha form.

$$\alpha(T) = T_r^{C_3(C_2-1)} \exp\left[C_1\left(1-T_r^{C_2C_3}\right)\right]$$

- This 3 parameter form eliminates the 2 problems with the Soave alpha for defined components

Mixing Rules

- The accuracy of correlating vapor-liquid equilibrium data using a cubic equation of state can be improved by choosing an appropriate mixing rule for calculating a and b for mixture.

- Expressions for mixing rules a and b are:

$$a = \sum_i \sum_j x_i x_j a_{ij} \qquad b = \sum_i x_i b_i$$

Mixing Rules for ‘a’

- Van der Waals Mixing Rule:

$$a_{ij} = \sqrt{a_i a_j}\left(1 - k_{ij}\right)$$

If $k_{ij} = k_{ji}$ …

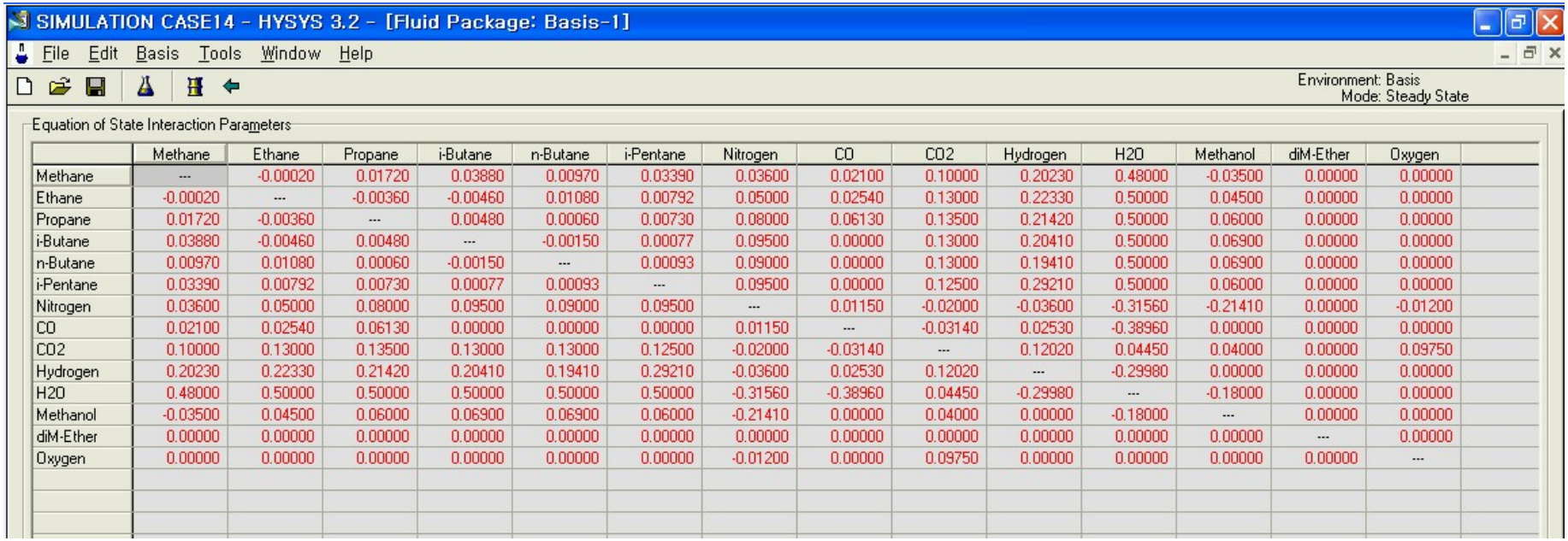
- Stryjek and Vera (1986) Mixing Rule:

$$a_{ij} = \sqrt{a_i a_j}\left[1 - \frac{k_{ij}k_{ji}}{x_i k_{ij} + x_j k_{ji}}\right]$$

- Panagiotopoulos (1985) Mixing Rule:

$$a_{ij} = \sqrt{a_i a_j}\left[1 - k_{ij} + \left(k_{ij} - k_{ji}\right)x_i\right]$$

PRSV EOS in HYSYS Simulator



Equation of State Interaction Parameters

	Methane	Ethane	Propane	i-Butane	n-Butane	i-Pentane	Nitrogen	CO	CO2	Hydrogen	H2O	Methanol	diM-Ether	Oxygen
Methane	---	-0.00020	0.01720	0.03880	0.00970	0.03390	0.03600	0.02100	0.10000	0.20230	0.48000	-0.03500	0.00000	0.00000
Ethane	-0.00020	---	-0.00360	-0.00460	0.01080	0.00792	0.05000	0.02540	0.13000	0.22330	0.50000	0.04500	0.00000	0.00000
Propane	0.01720	-0.00360	---	0.00480	0.00060	0.00730	0.08000	0.06130	0.13500	0.21420	0.50000	0.06000	0.00000	0.00000
i-Butane	0.03880	-0.00460	0.00480	---	-0.00150	0.00077	0.09500	0.00000	0.13000	0.20410	0.50000	0.06900	0.00000	0.00000
n-Butane	0.00970	0.01080	0.00060	-0.00150	---	0.00093	0.09000	0.00000	0.13000	0.19410	0.50000	0.06900	0.00000	0.00000
i-Pentane	0.03390	0.00792	0.00730	0.00077	0.00093	---	0.09500	0.00000	0.12500	0.29210	0.50000	0.06000	0.00000	0.00000
Nitrogen	0.03600	0.05000	0.08000	0.09500	0.09000	0.09500	---	0.01150	-0.02000	-0.03600	-0.31560	-0.21410	0.00000	-0.01200
CO	0.02100	0.02540	0.06130	0.00000	0.00000	0.00000	0.01150	---	-0.03140	0.02530	-0.38960	0.00000	0.00000	0.00000
CO2	0.10000	0.13000	0.13500	0.13000	0.13000	0.12500	-0.02000	-0.03140	---	0.12020	0.04450	0.04000	0.00000	0.09750
Hydrogen	0.20230	0.22330	0.21420	0.20410	0.19410	0.29210	-0.03600	0.02530	0.12020	---	-0.29980	0.00000	0.00000	0.00000
H2O	0.48000	0.50000	0.50000	0.50000	0.50000	0.50000	-0.31560	-0.38960	0.04450	-0.29980	---	-0.18000	0.00000	0.00000
Methanol	-0.03500	0.04500	0.06000	0.06900	0.06900	0.06000	-0.21410	0.00000	0.04000	0.00000	-0.18000	---	0.00000	0.00000
diM-Ether	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	---	0.00000
Oxygen	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	-0.01200	0.00000	0.09750	0.00000	0.00000	0.00000	0.00000	---

- Number of components: 14
- Required pairs of BIP's: $\frac{14 \times 13}{2!} = 91$

- Available pairs of BIP's built in HYSYS: 63 (69%)
- PRSV mixing rules have 2 adjustable parameters but….

Henry's Constant Built-In PRO/II Library

HENRY COEFFICIENTS FOR SET 'NRTL01'

LN(H) = C1 + C2/T + C3*LN(T) + C4*P
TEMPERATURE IN K
PRESSURE IN KPA

HENRY COEFFICIENTS

SOLUTE	SOLVENT	C1	C2	C3	C4	FROM
1	2	166.6723	-7847.1602	-21.8994	9.8692E-07	SIMSCI BANK
1	**5**	**N/A**	**N/A**	**N/A**	**N/A**	**VAPOR PRESSURE**
1	**6**	**N/A**	**N/A**	**N/A**	**N/A**	**VAPOR PRESSURE**
3	2	154.9483	-8498.7197	-20.0841	7.3032E-06	SIMSCI BANK
3	**5**	**217.0283**	**-10620.0000**	**-30.1360**	**0.0000**	**SIMSCI BANK**
3	6	134.5983	-6535.3999	-18.2570	0.0000	SIMSCI BANK
4	2	116.4133	-4881.3198	-14.7884	3.5529E-06	SIMSCI BANK
4	5	15.5643	271.3300	-0.5485	0.0000	SIMSCI BANK
4	6	223.4183	-8787.0996	-31.9150	0.0000	SIMSCI BANK
7	2	169.8923	-8132.2300	-22.3559	1.4409E-06	SIMSCI BANK
7	5	-63.1517	2617.0000	11.5200	0.0000	SIMSCI BANK
7	**6**	**N/A**	**N/A**	**N/A**	**N/A**	**VAPOR PRESSURE**
8	2	158.2643	-7260.1401	-20.7005	1.4409E-06	SIMSCI BANK
8	5	27.7643	-598.8800	-2.2701	0.0000	SIMSCI BANK
8	6	225.6083	-9275.2002	-32.1410	0.0000	SIMSCI BANK
9	2	155.5533	-7442.2900	-20.2359	1.6186E-06	SIMSCI BANK
9	5	15.2813	-185.0300	-0.3951	0.0000	SIMSCI BANK
9	6	195.2183	-8362.7998	-27.4340	0.0000	SIMSCI BANK

NRTL BIP's Built-In PRO/II Library

VLE LIQUID INTERACTION PARAMETERS FOR SET 'NRTL01'

NRTL BINARY COEFFICIENTS

I	J	A(I,J) / A(J,I)	B(I,J) / B(J,I)	C(I,J) / C(J,I)	ALPHAC / ALPHAT	UNITS FROM
2	5	0.511068	199.8540	0.00	0.2442	DEG K SIMSCI VLEBANK
		0.736107	-360.6920	0.00	0.0000	
2	6	12.300150	-4019.6919	330053.91	0.2000	DEG K SIMSCI VLEBANK
		-5.168679	2145.8921	74425.09	0.0000	
5	6	0.000000	653.0060	0.00	0.2951	DEG K SIMSCI VLEBANK
		0.000000	-18.9372	0.00	0.0000	

2: H2O
5: Methanol
6: DME

PR BIP's Built-In PRO/II Library

PR INTERACTION PARAMETERS

KIJ = A(I,J) + B(I,J)/T + C(I,J)/T**2

I	J	KA(I,J)	KB(I,J)	KC(I,J)	UNITS	FROM
1	2	0.2000	0.00	0.00	DEG K	SIMSCI BANK
1	3	-0.0300	0.00	0.00	DEG K	SIMSCI BANK
1	4	0.0900	0.00	0.00	DEG K	SIMSCI BANK
1	7	0.0300	0.00	0.00	DEG K	SIMSCI BANK
1	8	0.0120	0.00	0.00	DEG K	SIMSCI BANK
2	3	0.2100	0.00	0.00	DEG K	SIMSCI BANK
2	4	0.5630	0.00	0.00	DEG K	SIMSCI BANK
2	5	-0.0789	0.00	0.00	DEG K	SIMSCI BANK
2	7	0.5000	0.00	0.00	DEG K	SIMSCI BANK
2	8	0.5080	0.00	0.00	DEG K	SIMSCI BANK
3	4	-0.1622	0.00	0.00	DEG K	SIMSCI BANK
3	**5**	**0.0421**	**0.00**	**0.00**	**DEG K**	**SIMSCI BANK, 3: CO2, 5: Methanol**
3	7	0.0919	0.00	0.00	DEG K	SIMSCI BANK
3	8	-0.0170	0.00	0.00	DEG K	SIMSCI BANK
4	7	0.0160	0.00	0.00	DEG K	SIMSCI BANK
4	8	-0.0300	0.00	0.00	DEG K	SIMSCI BANK
5	8	-0.2700	0.00	0.00	DEG K	SIMSCI BANK
7	8	0.0350	0.00	0.00	DEG K	SIMSCI BANK
7	9	0.0500	0.00	0.00	DEG K	MW CORRELATION
8	9	-0.0119	0.00	0.00	DEG K	SIMSCI BANK

Coefficients in PR Alpha Form

PR PURE COMPONENT DATA

COMP	CRITICAL TEMPERATURE DEG C	CRITICAL PRESSURE KPA	ALPHA TYPE	C1	C2	C3
1	-140.20	3495.71	6	0.0746	0.8722	2.2635
2	374.20	22119.20	6	0.3846	0.8700	1.9637
3	31.04	7381.52	5	0.2047	0.8197	N/A
4	-239.90	1296.96	6	0.9267	4.2324	0.1200
5	239.43	8095.87	6	0.7515	0.9320	1.6042
6	126.90	5370.22	6	0.1140	0.8996	3.7232
7	-82.60	4600.15	5	0.1195	0.9040	N/A
8	-146.90	3394.39	6	0.5764	0.9093	0.6765
9	-118.40	5076.38	6	1.8858	3.7551	0.0922

Column Summary

UNIT 1, 'T01', 'CL-401'

TOTAL NUMBER OF ITERATIONS

IN/OUT METHOD 13

COLUMN SUMMARY

TRAY	TEMP DEG C	PRESSURE KPA	NET FLOW RATES KG-MOL/HR LIQUID	VAPOR	FEED	PRODUCT	HEATER DUTIES M*KCAL/HR
1	30.1	6000.00	341.2		329.8L	60.6V	
2	30.2	6000.00	342.8	71.9			
3	30.3	6000.00	343.7	73.6			
4	30.4	6000.00	344.7	74.4			
5	30.5	6000.00	346.1	75.4			
6	30.6	6000.00	348.3	76.9			
7	30.7	6000.00	351.4	79.1			
8	31.0	6000.00	355.4	82.1			
9	30.8	6000.00	360.8	86.2			
10	31.3	6000.00		91.6	97.0M	366.2L	

Stream Summary

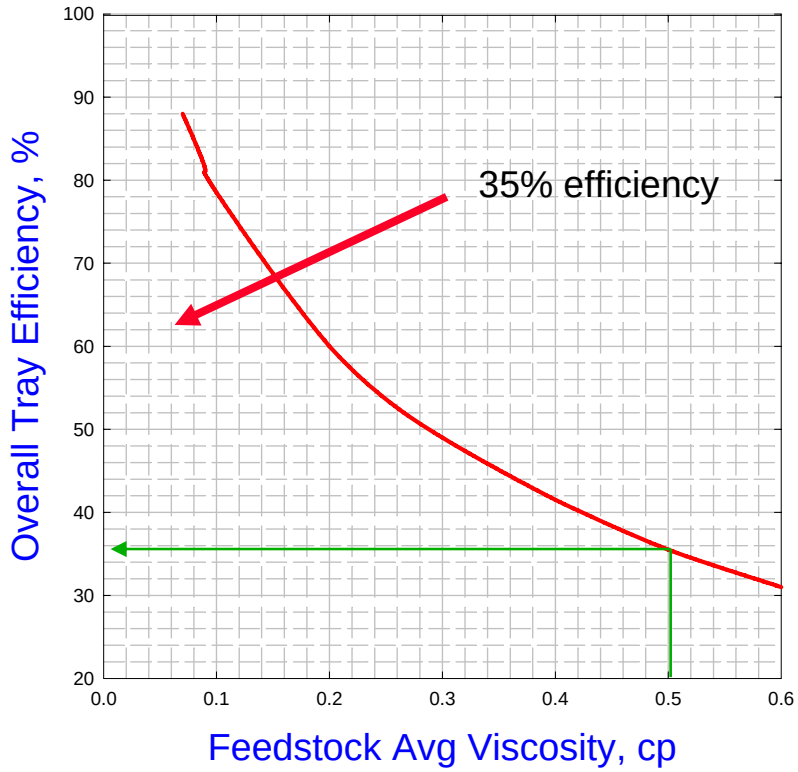
STREAM ID		18	19	20	21
	NAME				
	PHASE	MIXED	LIQUID	VAPOR	LIQUID
FLUID RATES, KG-MOL/HR					
1	CO	33.6217	0.0000	24.9896	8.6320
2	H2O	0.1842	0.0000	0.0000	0.1842
3	CO2	26.5713	0.0000	0.1687	26.4026
4	H2	36.3018	0.0000	34.8680	1.4338
5	METHANOL	0.0000	329.8000	0.3186	329.4814
6	DME	0.0000	0.0000	0.0000	0.0000
7	METHANE	0.1842	0.0000	0.1297	0.0545
8	N2	0.1067	0.0000	0.0997	6.9778E-03
9	O2	0.0000	0.0000	0.0000	0.0000
TOTAL RATE, KG-MOL/HR		96.9700	329.8000	60.5744	366.1955
TEMPERATURE, C		30.0000	30.0000	30.1297	31.3295
PRESSURE, KPA		6000.0000	6000.0000	6000.0000	6000.0000
ENTHALPY, M*KCAL/HR		0.0650	0.1858	0.0188	0.2320
MOLECULAR WEIGHT		22.6217	32.0420	13.0878	32.6828
MOLE FRAC VAPOR		0.9991	0.0000	1.0000	0.0000
MOLE FRAC LIQUID		9.3001E-04	1.0000	0.0000	1.0000

Overall Tray Efficiencies: Method 1

Method of Drickamer & Bradford [1]

Notes:

1) Based on 54 refinery columns.
2) Viscosity is average of feed as liquid at top & bottom temperatures of the column.
3) For Absorbers, use rich oil at exit temperature.
4) Efficiency is for key components.

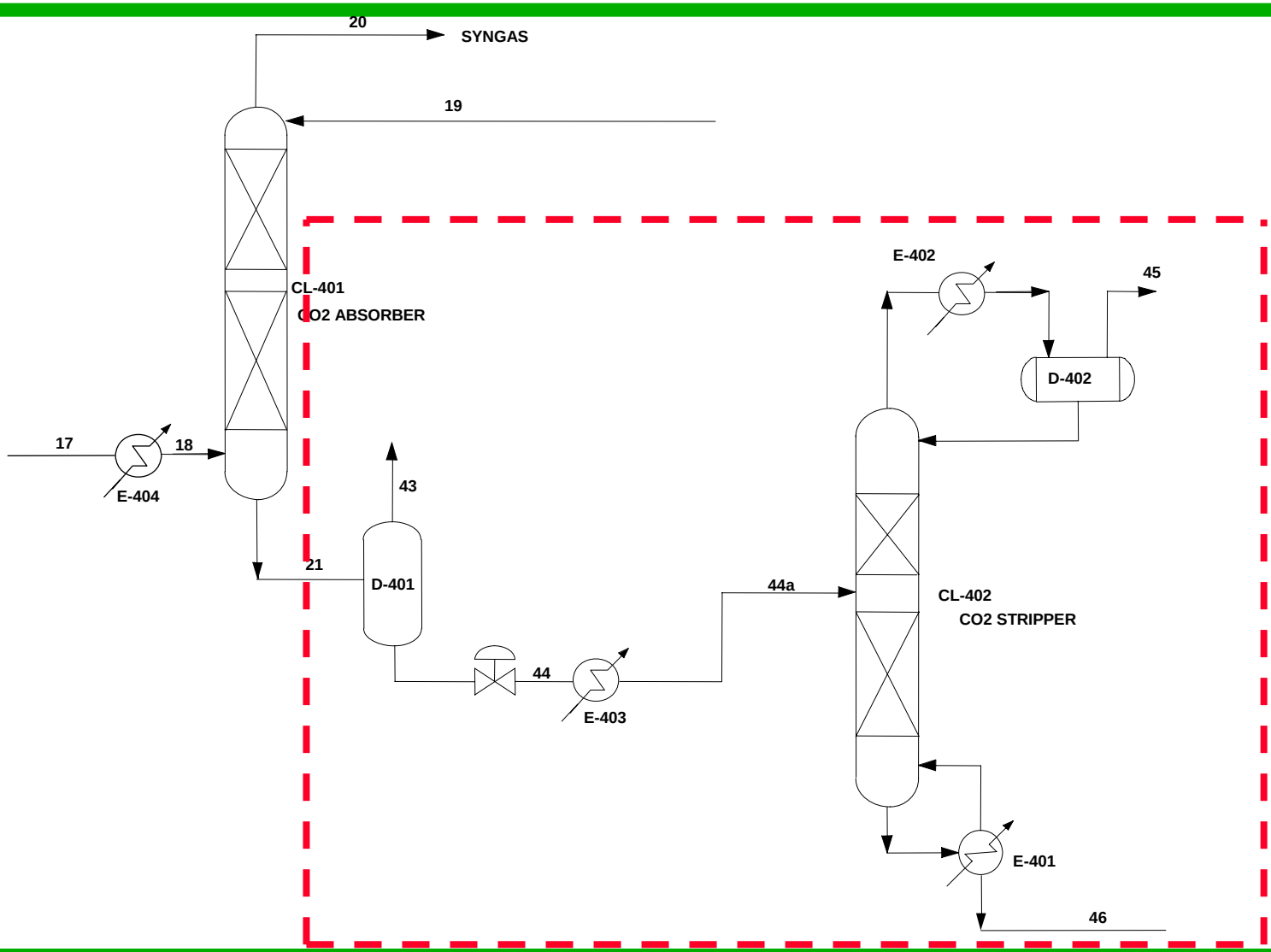


1. Tran. Am. Inst. Chem. Engrs, 39, 319 (1943).

Feedstock Average Viscosity

STREAM ID	18	**19**	20	21
NAME				
PHASE	VAPOR	LIQUID	VAPOR	LIQUID
-------- LIQUID --------				
RATE, KG-MOL/HR	N/A	329.767	N/A	365.809
K*KG/HR	N/A	10.566	N/A	11.956
M3/HR	N/A	13.533	N/A	16.817
GAL/MIN	N/A	59.582	N/A	74.041
STD LIQ RATE, M3/HR	N/A	13.281	N/A	15.011
SPECIFIC GRAVITY (H2O=1.0)	N/A	0.7964	N/A	0.7973
MOLECULAR WEIGHT	N/A	32.042	N/A	32.685
ENTHALPY, KCAL/KG	N/A	18.801	N/A	20.688
CP, KCAL/KG-C	N/A	0.613	N/A	0.595
DENSITY, KG/M3	N/A	780.811	N/A	710.988
Z (FROM DENSITY)	N/A	0.0970	N/A	0.1082
SURFACE TENSION, DYNE/CM	N/A	21.8405	N/A	19.5827
THERMAL COND, KCAL/HR-M-C	N/A	0.16968	N/A	0.14250
VISCOSITY, CP	N/A	**0.50292**	N/A	0.40460

DA-402 (CO2 Stripper)



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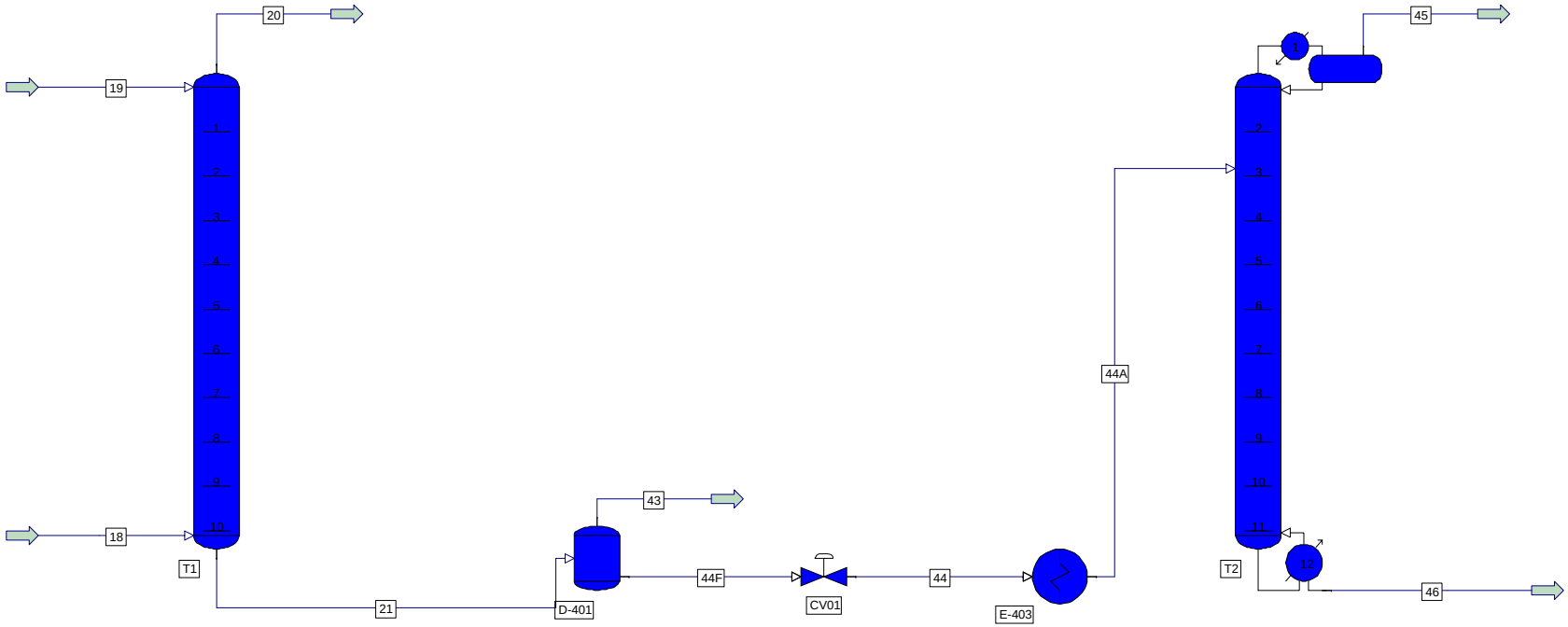
CO2 Stripper Simulation

- The CO_2 Stripper CL-402 serves to remove the absorb CO_2 from the methanol.
- The CO_2 is taken from overhead in the vapor phase and the methanol is taken from the bottom as a liquid.

Material Balance Around CO2 Stripper

	17	18	19	20	21	44	45	46
Temperature (°C)	110.8	30.0	30.0	29.54	33.1	33.82	31.29	185.7
Pressure (kPa, abs)	6,000	6,000	6,000	6,000	6,000	3,000	3,000	3,000
Flow (k-mole/hr)	**96.9700**	**96.9700**	**329.8000**	**67.8100**	**358.9000**	**356.5**	**27.1**	**329.5**
Total Kg/hr	**2,194.0**	**2,194.0**	**10,570.0**	**984.0**	**11,780.0**	**11,710.0**	**1,159.0**	**10,550.0**
Component Molar Flow								
1. CO	33.6200	33.6200	0.0000	32.3000	1.5800	0.7490	0.7520	0.0000
2. H2O	0.1842	0.1842	0.0000	0.0000	0.1795	0.1783	0.0000	0.1977
3. CO2	26.5700	26.5700	0.0000	0.0610	26.5227	26.5700	25.6600	0.0000
4. H2	36.300	36.300	0.0000	35.2300	1.0800	0.3922	0.3760	0.0000
5. Methanol	0.0000	0.0000	329.8000	0.2577	329.4702	329.5000	0.2000	329.3000
6. DME	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
7. CH4	0.1842	0.1842	0.0000	0.1356	0.0359	0.0357	0.0433	0.0000
8. N2	0.1067	0.1067	0.0000	0.0949	0.0000	0.0000	0.008115	0.0000
9. O2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000

Flow Sheet Drawing Using PRO/II



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CO2 Stripper Column Simulation

- Consider the following CO2 Stripper column to strip CO2 as a top product by obtaining methanol product as bottom.
  - Feed: CO2 Absorber Bottom Stream (Refer to feedstock characterization)
  - CO2 Stripper Column

    1) Number of Theoretical Stages: 12

    3) Overall Tray Efficiencies: Can by estimated by correlation

    4) Feed Tray Location: 3

Column Summary

COLUMN SUMMARY

TRAY	TEMP DEG C	PRESSURE KPA	NET FLOW RATES LIQUID KG-MOL/HR	NET FLOW RATES VAPOR KG-MOL/HR	NET FLOW RATES FEED KG-MOL/HR	NET FLOW RATES PRODUCT KG-MOL/HR	HEATER DUTIES M*KCAL/HR
1C	32.0	3200.00	72.0			36.0V	-0.6893
2	160.5	3200.00	102.9	108.0			
3	168.9	3200.00	566.7	138.8	365.7M		
4	180.5	3200.00	591.9	237.0			
5	183.3	3200.00	593.9	262.2			
6	184.9	3200.00	594.3	264.1			
7	185.9	3200.00	594.4	264.6			
8	186.6	3200.00	594.3	264.6			
9	187.1	3200.00	594.3	264.6			
10	187.5	3200.00	594.2	264.6			
11	187.8	3200.00	594.2	264.5			
12R	188.0	3200.00		264.4		329.7L	1.4342

SPECIFICATIONS

SPECIFICATION NUMBER	PARAMETER TYPE	TRAY NO	COMP NO	SPECIFICATION TYPE	SPECIFIED VALUE	CALCULATED VALUE
1 (ACTIVE)	TRAY LIQ	1		TEMPERATURE	3.200E+01	3.200E+01
2 (ACTIVE)	UNIT T2	1		MOL RRATIO	2.000E+00	2.000E+00

Stream Summary

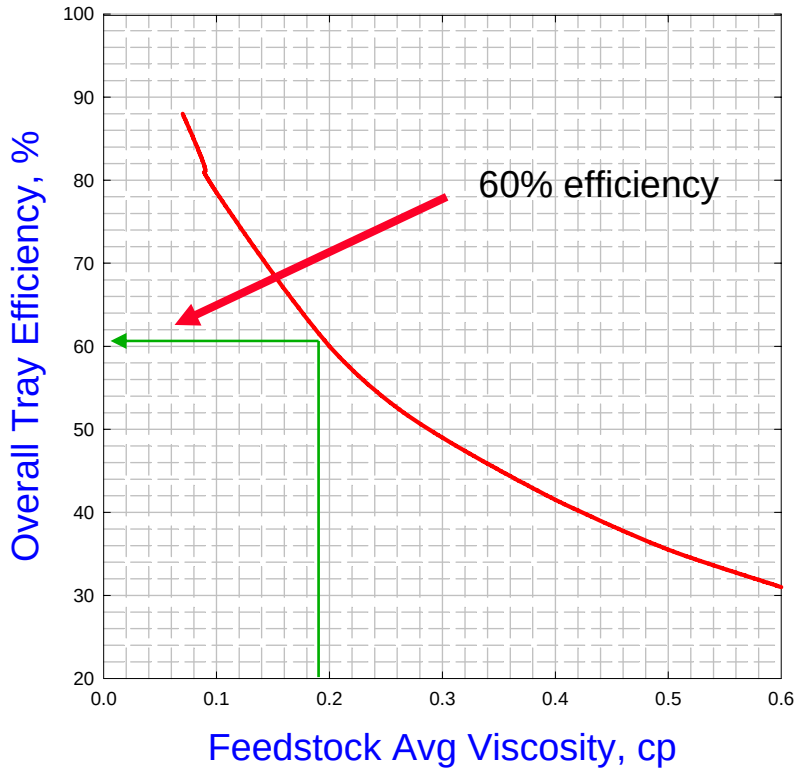
STREAM ID		45	46
	NAME		
	PHASE	VAPOR	LIQUID
FLUID RATES, KG-MOL/HR			
1	CO	8.3406	1.1364E-14
2	H2O	2.3033E-05	0.1842
3	CO2	25.7937	0.5275
4	H2	1.3935	1.5020E-17
5	METHANOL	0.4033	329.0119
6	DME	0.0000	0.0000
7	METHANE	0.0526	1.9869E-20
8	N2	6.7949E-03	8.7680E-19
9	O2	0.0000	0.0000
TOTAL RATE, KG-MOL/HR		35.9905	329.7237
TEMPERATURE, C		31.9998	187.9557
PRESSURE, KPA		3200.0000	3200.0000
ENTHALPY, M*KCAL/HR		0.0410	1.4860
MOLECULAR WEIGHT		38.4983	32.0533
MOLE FRAC VAPOR		1.0000	0.0000
MOLE FRAC LIQUID		0.0000	1.0000

Overall Tray Efficiencies: Method 1

Method of Drickamer & Bradford [1]

Notes:

1) Based on 54 refinery columns.
2) Viscosity is average of feed as liquid at top & bottom temperatures of the column.
3) For Absorbers, use rich oil at exit temperature.
4) Efficiency is for key components.



1. Tran. Am. Inst. Chem. Engrs, 39, 319 (1943).

Feedstock Average Viscosity

STREAM ID	43	44	**44A**	44F
NAME				
PHASE	VAPOR	MIXED	MIXED	LIQUID
-------- LIQUID --------				
RATE, KG-MOL/HR	N/A	358.677	349.880	365.714
K*KG/HR	N/A	11.742	11.426	11.954
M3/HR	N/A	16.221	16.821	16.809
GAL/MIN	N/A	71.418	74.061	74.009
STD LIQ RATE, M3/HR	N/A	14.717	14.325	15.007
SPECIFIC GRAVITY (H2O=1.0)	N/A	0.7986	0.7984	0.7973
MOLECULAR WEIGHT	N/A	32.737	32.657	32.688
ENTHALPY, KCAL/KG	N/A	20.551	64.351	20.687
CP, KCAL/KG-C	N/A	0.597	0.723	0.595
DENSITY, KG/M3	N/A	723.876	679.256	711.171
Z (FROM DENSITY)	N/A	0.0568	0.0496	0.1069
SURFACE TENSION, DYNE/CM	N/A	19.9752	14.5609	19.5878
THERMAL COND, KCAL/HR-M-C	N/A	0.14737	0.14240	0.14255
VISCOSITY, CP	N/A	0.42219	**0.18846**	0.40485

The End….